

(19)



Europäisches Patentamt
European Patent Office
Office européen des brevets



(51) Publication number:

0 564 678 A1

(12)

EUROPEAN PATENT APPLICATION(21) Application number: **92106034.9**(51) Int. Cl.⁵: **C10G 11/18**(22) Date of filing: **08.04.92**(43) Date of publication of application:
13.10.93 Bulletin 93/41(64) Designated Contracting States:
BE DE ES FR GB IT NL(71) Applicant: **UOP**
25 East Algonquin Road
Des Plaines, Illinois 60017-5017(US)(72) Inventor: **Lomas, David A.**
6 South Forest
Arlington Heights, Ill. 60005(US)
Inventor: **Haun, Edward C.**
340 Boulder Drive
Glendale Heights, Ill. 60139(US)
Inventor: **Sechrist, Paul A.**
281 Cornell Street
Des Plaines, Ill. 60016(US)(74) Representative: **Weber, Dieter, Dr. et al**
Weber, Dieter, Dr.,
Seiffert, Klaus, Dipl.-Phys.,
Lieke, Winfried, Dr., Postfach 61 45,
Gustav-Freytag-Strasse 25
D-65051 Wiesbaden (DE)(54) **FCC process and apparatus having a low volume dilute phase disengagement zone in the reaction vessel.**

(57) The operation of an FCC process and apparatus is improved by providing a low volume dilute disengagement zone in a reactor vessel containing a vented riser that provides an open discharge of catalyst and gaseous products into the upper region of the reactor vessel. In the improved system, the interior of the reactor vessel is arranged such that the outlet of the reactor riser is located close to and directed at the top of the reactor vessel, the reactor vessel operates with a dense bed of catalyst having an upper bed level that is only a short distance below the outlet of the reactor riser and the cyclone separators are located to the outside of the reactor riser and circulate catalyst back to the dense bed of the reactor section.

EP 0 564 678 A1

FIELD OF THE INVENTION

This invention relates generally to processes for the fluidized catalytic cracking (FCC) of heavy hydrocarbon streams such as vacuum gas oil and reduced crudes. This invention relates more specifically to a method for reacting hydrocarbons in an FCC reactor and separating reaction products from the catalyst used therein.

BACKGROUND OF THE INVENTION

The fluidized catalytic cracking of hydrocarbons is the main stay process for the production of gasoline and light hydrocarbon products from heavy hydrocarbon charge stocks such as vacuum gas oils or residual feeds. Large hydrocarbon molecules, associated with the heavy hydrocarbon feed, are cracked to break the large hydrocarbon chains thereby producing lighter hydrocarbons. These lighter hydrocarbons are recovered as product and can be used directly or further processed to raise the octane barrel yield relative to the heavy hydrocarbon feed.

The basic equipment or apparatus for the fluidized catalytic cracking of hydrocarbons has been in existence since the early 1940's. The basic components of the FCC process include a reactor, a regenerator and a catalyst stripper. The reactor includes a contact zone where the hydrocarbon feed is contacted with a particulate catalyst and a separation zone where product vapors from the cracking reaction are separated from the catalyst. Further product separation takes place in a catalyst stripper that receives catalyst from the separation zone and removes entrained hydrocarbons from the catalyst by counter-current contact with steam or another stripping medium. The FCC process is carried out by contacting the starting material whether it be vacuum gas oil, reduced crude, or another source of relatively high boiling hydrocarbons with a catalyst made up of a finely divided or particulate solid material. The catalyst is transported like a fluid by passing gas or vapor through it at sufficient velocity to produce a desired regime of fluid transport. Contact of the oil with the fluidized material catalyzes the cracking reaction. During the cracking reaction, coke will be deposited on the catalyst. Coke is comprised of hydrogen and carbon and can include other materials in trace quantities such as sulfur and metals that enter the process with the starting material. Coke interferes with the catalytic activity of the catalyst by blocking active sites on the catalyst surface where the cracking reactions take place. Catalyst is traditionally transferred from the stripper to a regenerator for purposes of removing the coke by oxidation with an oxygen-containing gas. An inventory of

catalyst having a reduced coke content, relative to the catalyst in the stripper, hereinafter referred to as regenerated catalyst, is collected for return to the reaction zone. Oxidizing the coke from the catalyst surface releases a large amount of heat, a portion of which escapes the regenerator with gaseous products of coke oxidation generally referred to as flue gas. The balance of the heat leaves the regenerator with the regenerated catalyst. The fluidized catalyst is continuously circulated from the reaction zone to the regeneration zone and then again to the reaction zone. The fluidized catalyst, as well as providing a catalytic function, acts as a vehicle for the transfer of heat from zone to zone. Catalyst exiting the reaction zone is spoken of as being spent, i.e., partially deactivated by the deposition of coke upon the catalyst. Specific details of the various contact zones, regeneration zones, and stripping zones along with arrangements for conveying the catalyst between the various zones are well known to those skilled in the art.

In order to obtain catalyst flow between the reactor and regenerator, a certain pressure balance is needed between the two vessels. Catalyst transferred to the regenerator is driven by the pressure in the upper section of the reactor vessel and the head of catalyst in the flow path from the reactor vessel to a control valve that controls the amount of catalyst entering the regenerator. Catalyst transfer from the regenerator to the riser is similarly driven by the pressure in the regenerator vessel and the head of catalyst in the flow path between the regenerator vessel and the bottom of the riser. There is usually a 4-10 psi (27.5-68.9 kPa) difference in pressure between the bottom of the riser and the top of the reactor vessel. Increases in pressure in the regenerator result in a corresponding increase in the pressure in the reactor. Any pressure increase of the regenerator relative to the reactor will be limited by the catalyst head in the flow path of the catalyst from the reactor to the regenerator. This interdependence of pressure often does not provide the most favorable pressure conditions in both the reactor and the regenerator. Low pressures are generally favored in the riser. Low pressures in the riser improve volatilization of the hydrocarbons and product yields. On the other hand, high pressures are favored in the regenerator to increase the coke burning capacity and improve the kinetics for coke burning. However, it is not possible to raise regenerator/reactor pressure differentials beyond a limited range due to the elevation of the available static catalyst head in the reactor vessel and stripper.

The rate of conversion of the feedstock within the reaction zone is controlled by regulation of the temperature of the catalyst, activity of the catalyst,

quantity of the catalyst (i.e., catalyst to oil ratio) and contact time between the catalyst and feed-stock. The most common method of regulating the reaction temperature is by regulating the rate of circulation of catalyst from the regeneration zone to the reaction zone which simultaneously produces a variation in the catalyst to oil ratio as the reaction temperatures change. That is, if it is desired to increase the conversion rate, an increase in the rate of flow of circulating fluid catalyst from the regenerator to the reactor is effected.

The hydrocarbon product of the FCC reaction is recovered in vapor form and transferred to product recovery facilities. These facilities normally comprise a main column for cooling the hydrocarbon vapor from the reactor and recovering a series of heavy cracked products which usually include bottom materials, cycle oil, and heavy gasoline. Lighter materials from the main column enter a concentration section for further separation into additional product streams.

Improvements in the reduction of product losses and the control of regeneration temperatures have been achieved by providing multiple stages of catalyst stripping and by raising the temperature at which the catalyst particles are stripped of products and other combustible compounds. Both of these methods will increase the amount of low molecular weight products that are stripped from the catalyst and will reduce the quantity of combustible material in the regenerator. A variety of arrangements are known for providing multiple stages of stripping and heating the spent catalyst to raise the temperature of the stripping zone. With increasing frequency it is being proposed to raise the temperature of the stripping zone by mixing the spent catalyst with hot regenerated catalyst from the regeneration zone.

As the development of FCC units has advanced, temperatures within the reaction zone were gradually raised. It is now commonplace to employ temperatures of 525°C (975°F). At higher temperatures, there is generally a loss of gasoline components as these materials crack to lighter components by both catalytic and strictly thermal mechanisms. At 1025°F (550°C), it is typical to lose 1% on the potential gasoline yield due to gasoline components thermally cracking into lighter hydrocarbon gases. As temperatures increase, to say 1025°F (550°C), most feedstocks lose up to 6% or more of the gasoline yield due to thermal cracking of gasoline components.

One improvement to FCC units, that has reduced the product loss by thermal cracking, is the use of riser cracking. In riser cracking, regenerated catalyst and starting materials enter a pipe reactor and are transported upward by the expansion of the gases that result from the vaporization of the

hydrocarbons, and other fluidizing mediums if present, upon contact with the hot catalyst. Riser cracking provides good initial catalyst and oil contact and also allows the time of contact between the catalyst and oil to be more closely controlled by eliminating turbulence and backmixing that can vary the catalyst residence time. An average riser cracking zone today will have a catalyst to oil contact time of 1 to 5 seconds. A number of riser designs use a lift gas as a further means of providing a uniform catalyst flow. Lift gas is used to accelerate catalyst in a first section of the riser before introduction of the feed and thereby reduces the turbulence which can vary the contact time between the catalyst and hydrocarbons.

The benefits of using lift gas to pre-accelerate and condition regenerated catalyst in a riser type conversion zone are well known. Lift gas typically has a low concentration of heavy hydrocarbons, i.e. hydrocarbons having a molecular weight of C_3 or greater are avoided. In particular, highly reactive type species such as C_3 plus olefins are unsuitable for lift gas. Thus, lift gas streams comprising steam and light hydrocarbons are generally used.

Riser cracking whether with or without the use of lift gas has provided substantial benefits to the operation of the FCC unit. These can be summarized as a short contact time in the reactor riser to control the degree of cracking that takes place in the riser and improved mixing to give a more homogeneous mixture of catalyst and feed. A more complete distribution prevents different times for the contact between the catalyst and feed over the cross-section of the riser such that some of the feed contacts the catalyst for a longer time than other portions of the feed. Both the short contact time and a more uniform average contact time for all of the feed with the catalyst has allowed over-cracking to be controlled or eliminated in the reactor riser.

Unfortunately, much of what can be accomplished in the reactor riser in terms of uniformity of feed contact and controlled contact time can be lost when the catalyst is separated from the hydrocarbon vapors. As the catalyst and hydrocarbons are discharged from the riser, they must be separated. In early riser cracking operations, the output from the riser was discharged into a large vessel. This vessel serves as a disengaging chamber and is still referred to as a reactor vessel, although most of the reaction takes place in the reactor riser. The reactor vessel has a large volume. Vapors that enter the reactor vessel are well mixed in the large volume and therefore have a wide residence time distribution that results in relatively long residence times for a significant portion of the product fraction. Product fractions that encounter extended residence times can undergo ad-

ditional catalytic and thermal cracking to less desirable lower molecular weight products.

In an effort to further control the contact time between catalyst and feed vapors, there has been continued investigation into the use of cyclones that are directly coupled (see U.S. Patent No. 4,737,346) to the end of the reactor riser. This direct coupling of cyclones to the riser provides a quick separation of most of the product vapors from the catalyst. Therefore, contact time for a large portion of the feed vapors can be closely controlled. One problem with directly coupling cyclones to outlet of the reactor riser is the need for a system that can handle pressure surges from the riser--see U.S. Patent No. 4,624,772 for an example. These pressure surges and the resulting transient increase in catalyst loading inside the cyclones can overload the cyclones such that an unacceptable amount of fine catalyst particles are carried over with the reactor vapor into downstream separation facilities. Therefore, a number of apparatus arrangements have been proposed for direct coupled cyclones that significantly complicate the arrangement and apparatus for the direct coupled cyclones, and either provide an arrangement where a significant amount of reactor vapor can enter the open volume of the reactor/vessel or compromise the satisfactory operation of the cyclone system by subjecting it to the possibility of temporary catalyst overloads.

Although direct coupled cyclone systems can help to control contact time between catalyst and feed vapors, they will not completely eliminate the presence of hydrocarbon vapors in the open space of a reactor vessel. Product vapors are still present in this open space from the stripped hydrocarbon vapors that are removed from the catalyst and pass upwardly into an open space above the stripping zone. The amount of hydrocarbon vapors is also increased by direct coupled cyclone arrangements that allow feed vapors to enter the open space that houses the cyclones. Since the dilute phase volume of the reactor vessel remains unchanged when direct coupled cyclones are used and less hydrocarbon vapors enter the dilute phase volume from the risers, the hydrocarbon vapors that do enter the dilute phase volume will be there for much longer periods of time when a direct coupled cyclone system is used. (The terms "dense phase" and "dilute phase" catalysts as used in this application are meant to refer to the density of the catalyst in a particular zone. The term "dilute phase" generally refers to a catalyst density of less than 20 lbs/ft³ (480 kg/m³) and the term "dense phase" refers to catalyst densities above 30 lbs/ft³ (480 kg/m³). Catalyst densities in the range of 20 to 30 lbs/ft³ (320 to 480 kg/m³) can be considered either dense or dilute depending on the density of

the catalyst in adjacent zones or regions.) In other words, when a direct coupled cyclone system is used, less product vapors may enter the open space of the reactor vessel, but these vapors will have a much longer residence time in the reactor vessel. As a result, any feed components left in the reactor vessel are substantially lost to overcracking.

The very low gas flow rate through the reactor vessel can also promote coke deposition on the interior of the vessel. The long residence time of heavy hydrocarbons at relatively high temperature in the upper section of the reactor vessel promotes the formation of coke. These coke deposits interfere with the function of the reactor vessel by forming thick deposits on the interior of the vessel thereby insulating and locally cooling portions of the metal shell. Such locally cooled portions promote the condensation of corrosive materials that can damage the reactor vessel. In addition, other problems are created by the large coke deposits which can, from time to time, break off in large chunks and block the flow of catalyst through the vessels or conduits.

One apparatus that has been known to promote quick separation between the catalyst and the vapors in the reactor vessels is known as a ballistic separation device (see U.S. Patent No. 4,792,437) which is also referred to as a vented riser. The structure of the vented riser in its basic form consists of a straight portion of conduit at the end of the riser and an opening that is directed upwardly into the reactor vessel with a number of cyclone inlets surrounding the outer periphery of the riser near the open end. The apparatus functions by shooting the high momentum catalyst particles past the open end of the riser where the gas collection takes place. A quick separation between the gas and the vapors occurs due to the relatively low density of the gas which can quickly change directions and turn to enter the inlets near the periphery of the riser (see U.S. Patent No. 4,295,961) while the heavier catalyst particles continue along a straight trajectory that is imparted by the straight section of riser conduit. The vented riser has the advantage of eliminating any dead area in the reactor vessel where coke can form while providing a quick separation between the catalyst and the vapors. However, the vented riser still has the drawback of operating within a large open volume in the reactor vessel.

The problem addressed by the present invention is the need to obtain a quick separation between catalyst and product vapors leaving an FCC riser in a system that minimizes overcracking of product vapors and the carryover of fine catalyst particles with the product vapors.

BRIEF DESCRIPTION OF THE INVENTION

It is an object of this invention to reduce the hydrocarbon residence time in a reactor vessel.

It is another object of this invention to improve the operation of a vented riser in the separation of catalyst and hydrocarbon vapors.

A further object of this invention is to control the residence time in a reactor vessel section of an FCC reaction zone.

This invention is an FCC process that is arranged so that the outlet end of a reactor riser discharges into an upper portion of a reactor vessel that functions as an open disengagement zone and the discharge end of the reactor riser is located near the top of a dense catalyst bed in the reactor vessel. This arrangement is facilitated by putting the cyclones or gas solid separators on the outside of the reactor vessel. It has been discovered that when a vented riser is used in combination with externally located cyclones or separation devices the upper level of a dense catalyst bed can be maintained near the top of a reactor vessel. The high level of the dense catalyst bed relative to the riser outlet reduces the dilute phase volume in the disengagement vessel so that hydrocarbon residence time is reduced and the reactor vessel height required above the dense bed level is decreased.

The decreased height of the dilute phase in the reactor vessel offers a number of benefits. The decreased dilute phase height can shorten the reactor vessel and its overall elevation. Alternately, the overall elevation of the reactor vessel can be maintained and the additional height can be used to maintain a longer vertical length for the dense catalyst phase. Additional catalyst stripping can be performed in the additional length of dense catalyst bed. In addition, an increased height of dense catalyst increases the pressure drop between the reactor vessel and the regenerator control valve so that higher regenerator pressures can be maintained without raising the pressure in the reactor zone. The dense bed level is also susceptible to a substantial degree of variation so that the overall residence time of hydrocarbon vapors in the reactor vessel can be adjusted. This permits the use of very short residence time for certain feedstocks and an increase in residence time for more refractory feedstocks without a variation in the feed rate to the reactor riser. These benefits show that this invention will provide much of the same improvement offered by direct coupled cyclones in regard to reducing overcracking while giving a much greater degree of flexibility that is combined with the increased reliability of an open discharge type reactor riser.

Accordingly in one embodiment, this invention is a process for the fluidized catalytic cracking of an FCC feedstock. The process includes the steps of passing the FCC feedstock and catalyst particles to a reactor riser having an effective riser outlet diameter, transporting the catalyst and feedstock upwardly through the riser thereby converting the feedstock to a gaseous product stream, and producing spent catalyst particles by the deposition of coke. The mixture of spent catalyst particles and gaseous products are discharged into a reactor vessel in an upward direction from a discharge end of the riser located from 1 to 12 outlet riser diameters below the upper end of the reactor vessel thereby providing an initial separation of the spent catalyst from the gaseous products. The separated catalyst passes downward through the vessel where it is maintained as a dense catalyst bed to form a stripping zone while a stripping gas passes upwardly through the reactor vessel. The upper surface of the bed is a minimum distance of 2 feet (.6 m) or 1 riser diameter below the riser outlet end, and a maximum distance of from 8 feet (2.4 m) to 4 riser diameters below the outlet end of the riser. A mixture of gaseous product stripping fluid and spent catalyst particles is withdrawn from the reactor vessel and transferred to a particle separator that is located outside of the reactor vessel wherein the gaseous components are separated from the spent catalyst which is ultimately returned to the regeneration zone. Spent catalyst particles are passed from the stripping zone into a regeneration zone and contacted therein with a regeneration gas in the regeneration zone to combust coke from the catalyst particles and produce regenerated catalyst particles for transfer to the reactor riser.

This invention can also be described in the context of an apparatus. The apparatus includes an upwardly directed riser conduit having an upwardly directed outlet end having an outlet opening corresponding to an equivalent diameter, referred to herein as the "riser outlet diameter", a reactor vessel that surrounds the outlet end and has an upper end located 5 to 12 riser outlet diameters above the outlet end of the riser. Gas solids separation devices are located outside of the reactor vessel and these devices have an inlet, a gas outlet and a solids outlet. Means are provided for collecting a mixture of gas and catalyst from the upper half of the reactor vessel and communicating the mixture of catalyst and gas to the inlet of the separation device. Means are provided for returning catalyst particles from the collector to the reactor vessel, and means are provided for withdrawing catalyst from the bottom of the reactor vessel and transferring the catalyst to a regeneration vessel.

BRIEF DESCRIPTION OF THE DRAWINGS

The Figure shows a reactor/regenerator system for an FCC process arranged in accordance with this invention.

DETAILED DESCRIPTION OF THE INVENTION

This invention relates generally to the reactor side of the FCC process. This invention will be useful for most FCC processes that are used to crack light or heavy FCC feedstocks. The process and apparatus aspects of this invention can be used to modify the operation and arrangement of existing FCC units or in the design of newly constructed FCC units.

This invention uses the same general elements of many FCC units. A reactor riser provides the primary reaction zone. A reactor vessel and cyclone separators remove catalyst particles from the gaseous product vapors. A stripping zone removes a large percentage of sorbed catalyst particles from the surface of the catalyst. Spent catalyst from the stripping zone is regenerated in a regeneration zone having one or more stages of regeneration. Regenerated catalyst from the regeneration zone is used in the reactor riser. A number of different arrangements can be used for the elements of the reactor and regenerator sections. The description herein of specific reactor and regenerator components is not meant to limit this invention to those details except as specifically set forth in the claims.

An overview of the basic process operation can be best understood with reference to the drawing. Regenerated catalyst from a lower portion 12 of a regeneration vessel 10 is transferred by a conduit 14, at a rate regulated by a control valve 16, to a Y-section 18. Lift gas injected into the bottom of Y-section 18, by a conduit 20, carries the catalyst upward through a lower riser section 22. Feed is injected into the riser above lower riser section 22 by feed injection nozzles 24.

The mixture of feed, catalyst and lift gas travels up an intermediate section of the riser 26 and into an upper internal riser section 28 that terminates in an upwardly directed outlet end 30 that is located in a dilute phase region 32 of a reactor vessel 34. The gas and catalyst are separated in dilute phase section 32. Conduits 36 collect gas from the dilute phase section and transfer it to a collection chamber 38. From collection chamber 38, a T-type piping arrangement 40 distributes the gas which still contains a small amount of catalyst particles to a pair of cyclone separators 42. Relatively clean product vapors are recovered from the outlets of cyclones 42 by a manifold 44 and withdrawn from the process through an outlet 46. Catalyst separated by cyclone separators 42 is carried back to

reactor vessel 34 by dip pipe conduits 48. Spent catalyst from dilute phase section 32 and the dip pipe conduits form a dense catalyst bed 50 in a lower portion of the reactor vessel 34. Separated catalyst passes downwardly through the vessel at an average rate of less than 20 lb/ft²/sec (98 kg/m²/sec). The dense catalyst bed extends downward into a stripping vessel 52 that operates as a stripping zone. Stripping fluid enters a lower portion of the stripping vessel 52 through a distributor 54 and travels upward through the stripping vessel and reactor vessel in countercurrent flow to the downward moving catalyst. As the catalyst moves downward, it passes over reactor stripping baffles 56 and 58 and stripper baffles 60 and 62 and is transferred into an upper section 68 of the regenerator vessel 10 by a conduit 64 at a rate regulated by a valve 66 control. The catalyst particles are contacted with an oxygen-containing gas in upper section 68 of the regeneration zone. A distributor 70 receives the oxygen-containing gas from a conduit 71 and distributes the gas over the cross-section of the regeneration vessel. Regenerated catalyst is withdrawn from upper section 68 and transferred by a conduit 72 into lower portion 12 of the regeneration vessel at a rate regulated by a control valve 74. Catalyst in the lower portion 12 is contacted with additional regeneration gas that enters the vessel through conduit 76 and is distributed over the cross-section of the vessel by dome style distributor 78. Catalyst in lower section 12 is fully regenerated and withdrawn by conduit 14 in the manner previously described. The products of coke combustion in the lower regeneration section 12 rise upwardly and flow into the upper regenerator section 68 through a series of internal vents 80. Flue gas from lower section 12 is mixed with flue gas generated in upper section 68 and withdrawn through an inlet 82 of cyclones 84. The flue gas entering cyclone 84 contains a small amount of fine catalyst particles that are removed by the cyclones and returned to the regenerator by dip legs 86. Flue gas leaving the cyclones is collected in a chamber 88 that leaves the regenerator through a conduit 90.

The reactor riser of this invention is laid out to perform an initial separation between the catalyst and gaseous components in the riser. The term "gaseous components" includes lift gas, product gases and vapors, and unconverted feed components. The drawing shows this invention being used with a riser arrangement having a lift gas zone 22. It is not necessary that a lift gas zone be provided in the riser in order to enjoy the benefits of this invention. However, the end of the riser must terminate with one or more upwardly directed openings that discharge the catalyst and gaseous mixture in an upward direction into a dilute phase

section of the reactor vessel. The open end of the riser can be of an ordinary vented riser design as described in the prior art patents of this application or of any other configuration that provides a substantial separation of catalyst from gaseous material in the dilute phase section of the reactor vessel. It is believed to be important that the catalyst is discharged in an upward direction in order to minimize the distance between the outlet end of the riser and the top of the dense phase catalyst bed in the reactor vessel. The flow regime within the riser will influence the separation at the end of the riser. Typically, the catalyst circulation rate through the riser and the input of feed and any lift gas that enters the riser will produce a flowing density of between 3 lbs/ft³ to 20 lbs/ft³ (48 to 320 kg/m³) and an average velocity of about 10 ft/sec to 100 ft/sec (3 to 30 m/sec) for the catalyst and gaseous mixture. The length of the riser will usually be set to provide a residence time of between .5 to 5 seconds at these average flow velocity conditions.

The velocity at which the catalyst and gaseous mixtures discharge from end 30 of the riser also influences the placement of the end of the riser relative to the top of the reactor vessel. This distance indicated by the letter "A" in the drawing is set on the basis of the flow rate to riser. In the interest of minimizing the dilute volume of catalyst in the reactor vessel, distance "A" should be kept as short as possible. Nevertheless, there is need for some space between the end of the riser to avoid direct impingement and the resulting erosion of the top of the reactor vessel and to allow the discharge of catalyst from the end of the riser to provide a separation while preventing the reentrainment of catalyst particles that are separated by the initial discharge from the riser with the gas stream that is collected from the upper section of the reactor vessel. Since the reactor riser is usually designed for a narrow range of exit velocities between 20 to 100 ft/sec (6.1 to 30.5 m/sec), distance "A" can be set on the basis of riser outlet diameter. In order to avoid erosion of the upper surface of the reactor vessel and to promote the initial separation of the catalyst from the gaseous components, the distance "A" should be at least one outlet diameter or more. The avoidance of catalyst re-entrainment after discharge of the riser is influenced by both the riser velocity and the flowing density of the catalyst as it passes downwardly through the reactor vessel. For most practical ranges of catalyst density in the riser, the distance of 1.5 to 12 riser outlet diameters, and more preferably 1.5 to 6 riser diameters, for dimension "A" is adequate for a flowing catalyst density, often referred to as "catalyst flux", of about 50 - 200 lb/ft²/sec. (244 - 976 kg/m²/sec).

The total dilute phase volume in the reactor vessel is determined by the diameter of the reactor vessel, the distance from the end of the riser to the top of the reactor vessel, dimension "A", and the distance from the discharge end of the riser to the top of the dense bed level in the reactor vessel which is shown as dimension "B" in the Figure. In order to prevent re-entrainment of catalyst particles into the gases that are withdrawn from the reactor vessel, a minimum distance is required from the top of the reactor riser to the top of the catalyst bed level. This dimension is primarily influenced by the superficial velocity of the gases that flow upwardly through dense bed 50. In order to minimize the potential for re-entrainment of the gaseous compounds passing through bed 50, the superficial velocity is typically below 0.5 ft/sec (.15 m/sec). The gaseous components passing upward through bed 50 are made up of stripping fluid and hydrocarbons that are desorbed from the surface of the catalyst. The amount of stripping gas that enters the stripping vessel is usually proportional to the volume of voids in the catalyst. For most reasonable catalyst to oil ratios in the riser, the amount of stripping gas that must be added to displace the void volume of the catalyst will not exceed 6 wt% of the feed rate. Accordingly, the relative ratio of catalyst passing downwardly through the reactor vessel and the stripping fluid as well as other displaced hydrocarbons flowing upwardly through the reactor vessel will remain relatively constant. Thus, the primary variable in controlling the superficial gas velocity upward through the dense catalyst bed is the diameter of the reactor vessel. As long as the superficial velocity of the gases rising through bed 50 are kept in a range below 0.5 ft/sec (.15 m/sec) and preferably below about 0.1 ft/sec (.03 m/sec), a distance "B" of 2 to 8 feet (.6 to 2.4 m) will prevent re-entrainment of the catalyst and the gases that are leaving the reactor vessel. For most reactor risers, the 2 to 8 feet will equal approximately 1 to 4 riser outlet diameters.

The manner in which the gaseous vapors are withdrawn from the dilute phase volume of the reactor vessel will also influence the initial separation and the degree of re-entrainment that is obtained in the reactor vessel. In order to improve this disengagement and avoid re-entrainment, the Figure shows the use of an annular collector 92 that surrounds the end 30 of the riser. Collector 92 is supported from the top of the reactor vessel 34 by withdrawal conduits 36. Withdrawal conduits 36 are symmetrically spaced around the annular collector and communicate with the annular collector through a number of symmetrically spaced openings to obtain a balanced withdrawal of gaseous components around the entire circumference of the reactor riser. All of the stripping gas and gaseous

components from the reactor riser are withdrawn by annular collector 92 for the process arrangement shown in the Figure. All of the product gases from conduits 36 are transferred to the cyclones 42.

The Figure shows an arrangement for transferring gases from the conduits 36 to the cyclones that avoids a mal-distribution of the catalyst and gas mixture to the different cyclones. The simplest way to connect the gas conduits with the cyclones is to directly couple one conduit to a corresponding cyclone. This arrangement would also have the advantage of minimizing the flow path between the annular collector of the riser and the cyclones where the final separation of catalyst and gas is performed. However, for reasons unknown, it has been found that mixtures of catalyst and gas that are taken from the reactor through a series of conduits may preferentially flow to one conduit. The resulting heavier loading of catalyst and gas can overload the cyclone to which it is directed. For this reason, the Figure shows the use of a chamber 38 that commonly collects the gas from all cyclone conduits 36 and redistributes the gas to the individual cyclones. Although providing chamber 38 and T-section 40 increases the residence time for the catalyst and gas mixture as it flows from the reactor vessel to the cyclone inlets, this minor increase in residence time will not have a substantial impact on the quality of the product recovered from the cyclones. The avoidance of mal-distribution may also be accomplished by the use of a catalyst and gas separation device other than cyclones.

Catalyst recovered by the cyclones can be returned to the process at any convenient location. Whatever type of gas and catalyst separation device is utilized, the catalyst separated therefrom is returned to the process. The catalyst may be returned to any point of the process that puts it back into the circulating inventory of catalyst. The drawing shows the use of conventional cyclones with the dip legs 48 returning near the upper bed level 51 of dense bed region 50. Preferably, the catalyst will be returned to the dense bed in the reactor vessel or stripping vessel.

With the cyclones removed from the reactor vessel, the diameter of the reactor vessel is no longer affected by the need to provide adequate space for a separation device therein. Accordingly, the diameter of the reactor vessel can be set on the basis of the superficial velocity of gas passing upward through the dense bed and the catalyst flux entering the reactor vessel. The criteria for both of these parameters, as previously discussed, will permit the use of a smaller reactor diameter than has been found in the prior art. The smaller reactor vessel diameter further decreases the volume of

the dilute phase in the reactor vessel. When this invention is used with a new reactor vessel, the diameter can be kept low enough such that the average residence time in the dilute phase of the reactor vessel will be less than three seconds. Again, since the superficial velocity and catalyst flux are influenced by a well-known range of catalyst density and velocity conditions in the riser, the diameter of the reactor vessel when initially designed in accordance with this invention will preferably be between three and five times the outlet diameter of the riser. Alternately the dilute phase volume of the reactor vessel above the top of the catalyst bed can be kept to less than five times the volume of the reactor riser through which the feed passes.

Catalyst that is initially separated from the gaseous components as it enters the reactor vessel, passes downwardly through the vessels as previously described. As this catalyst progresses through the vessel, it preferably contacts a series of baffles that improve the contact of the catalyst with a stripping gas that passes upwardly through the vessel. In the embodiment of the invention shown in the Figure, the catalyst passes through a stripping section in the upper portion of the vessel and a separate stripping vessel located therebelow. The Figure shows the baffles 56 and 62 located on the exterior of the vessel walls and baffles 58 and 60 located down the length of the riser through the lower portion of the reactor vessel and the stripping vessel. These stripping baffles function in the usual manner to cascade catalyst from side to side as it passes through the vessel and increase the contact of the catalyst particles with the stripping steam as it passes upward in countercurrent contact with the catalyst. Dense bed 50 has a relatively long length in reactor vessel 34. There is no requirement for a long dense bed length in the reactor vessel and the dense bed length shown in the Figure stems from the type of arrangement depicted in the Figure. The Figure depicts a retrofit of this invention into an existing regeneration and reactor section where the tangent length of the reactor vessel was set by the previous arrangement that place the cyclone separators inside the reactor vessel. When the method of this invention is employed in the initial design of a reactor vessel, the tangent length can be substantially reduced so that upper bed level 51 is near the top of a stripper vessel.

Nevertheless, the height of the dense catalyst bed in the reactor 34 increases the total height of the dense phase catalyst above control valve 66. This additional height of dense bed catalyst can be used advantageously in the FCC process. First, the additional length of dense bed catalyst provides an elongated region for increased contact between the stripping fluid and the catalyst. Therefore, a greater

degree of stripping can be obtained by the extended length of the dense catalyst bed. In addition, the hydrostatic head of catalyst from the top surface 51 to control valve 66 produces a relatively high pressure drop between control valve 66 and bed level 51. This pressure drop can total 7 psi (48 kPa) or more. This additional pressure allows the regenerator to be operated at a higher pressure than the reactor section. As previously described, there are substantial benefits to operating the regeneration zone at higher pressures and the reaction zone at lower pressures.

The additional height of dense bed can also be used to incorporate a hot stripping section. The hot stripping section will utilize catalyst from the regeneration zone to supply heat to the stripping section and increase the desorption of hydrocarbons and volatile components from the surface of the catalyst. A suitable lift system can be used to transport the catalyst upward from the regeneration zone into a stripping zone at a desired elevation.

The catalyst is withdrawn from the stripping zone and transferred to a regeneration zone. Any well-known regenerator arrangement for removing coke from the catalyst particles by the oxidative combustion of coke and returning catalyst particles to the reactor riser can be used.

Claims

1. A process for the fluidized catalytic cracking (FCC) of an FCC feedstock said process comprising:

- a) passing said FCC feedstock and catalyst particles to a reactor riser having a riser outlet diameter and transporting said catalyst and feedstock upwardly through said riser thereby converting said feedstock to a gaseous products and producing spent catalyst particles by the deposition of coke on said catalyst particles;

- b) discharging a first mixture of spent catalyst particles and gaseous products in an upward direction into a reactor vessel from a discharge end of said riser located from 5 to 12 riser outlet diameters below the upper end of said reactor vessel thereby providing an initial separation of the spent catalyst from the gaseous products;

- c) passing separated catalyst downward through said reactor vessel;

- d) maintaining said separated catalyst in said reactor vessel as a dense catalyst bed to form a stripping zone and passing a stripping gas upward through the stripping zone;

- e) maintaining the upper surface of said dense catalyst bed a distance of between

one to four riser outlet diameters below said riser discharge;

- f) passing stripped catalyst from said stripping zone into a regeneration zone and contacting said spent catalyst with a regeneration gas in said regeneration zone to combust coke from said catalyst particles and produce regenerated catalyst particles for transfer to said reactor riser; and,

- g) withdrawing a second mixture of gaseous products, stripping fluid, and spent catalyst particles from said reactor vessel and transferring said mixture to a particle separator located outside of the reactor vessel and separating gaseous components from said spent catalyst particles in said separator zone.

2. The process of claim 1 wherein said first mixture is discharged from said riser at a velocity in a range of from 20 to 100 ft/sec. (6.1 to 30.5 m/sec).

3. The process of claim 1 wherein said second gas mixture is withdrawn from an annular collector that surrounds said outlet end of said riser and said particle separators comprise at least one cyclone.

4. The process of claim 1 wherein the gaseous products discharged in step (b) have an average residence time of less than three seconds in said reactor vessel.

5. The process of claim 1 wherein said reactor vessel has a diameter that is between three and five times the outlet diameter of said riser.

6. The process of claim 1 wherein the pressure drop in said reactor vessel from the bottom to the top of said dense bed is at least 7 psi (48.3 kPa).

7. The process of Claim 1 wherein separated catalyst passes downward through said reactor vessel at an average rate of less than 20 lb/ft²-sec (98 kg/m²/sec) and the stripping gas passes upwardly through the reactor vessel at an average superficial velocity of less than .5 ft/sec (.15 m/sec);

8. An apparatus for the fluidized catalytic cracking of an (FCC) feedstock, said apparatus comprising:

- a) an upwardly directed riser conduit having an upwardly directed outlet end having an equivalent outlet diameter:

- b) a reactor vessel surrounding said outlet end and having an upper end located 1 to 12 riser outlet diameters above said outlet end;
- c) a gas solids separation device located outside of said of said reactor vessel having an inlet, a gas outlet, and a solids outlet;
- d) means for collecting a mixture of gas and catalyst from an upper portion of said reactor vessel and communicating said mixture of catalyst and gas to the inlet of said separation device;
- e) means for introducing a gaseous stripping medium into the reactor vessel below the outlet end; and
- f) means for returning catalyst particles from said collector and from the bottom of the disengaging vessel to said reactor vessel.
9. The apparatus of claim 8 wherein said gas solid separation device comprises at least one cyclone separator located outside of said reactor vessel and said means for collecting a mixture of gas and catalyst includes an annular collector that surrounds the outlet end of said riser.
10. The apparatus of claim 8 wherein a dip leg communicates catalyst from each catalyst outlet to said reactor vessel and the outlet of each dip leg is located below said riser outlet.
11. The process of claim 1 wherein the dilute phase volume of said disengaging zone above the top of said catalyst bed is less than 5 times the volume of said reactor riser between the point where the feed enters the riser and said discharge end.

5

10

15

20

25

30

35

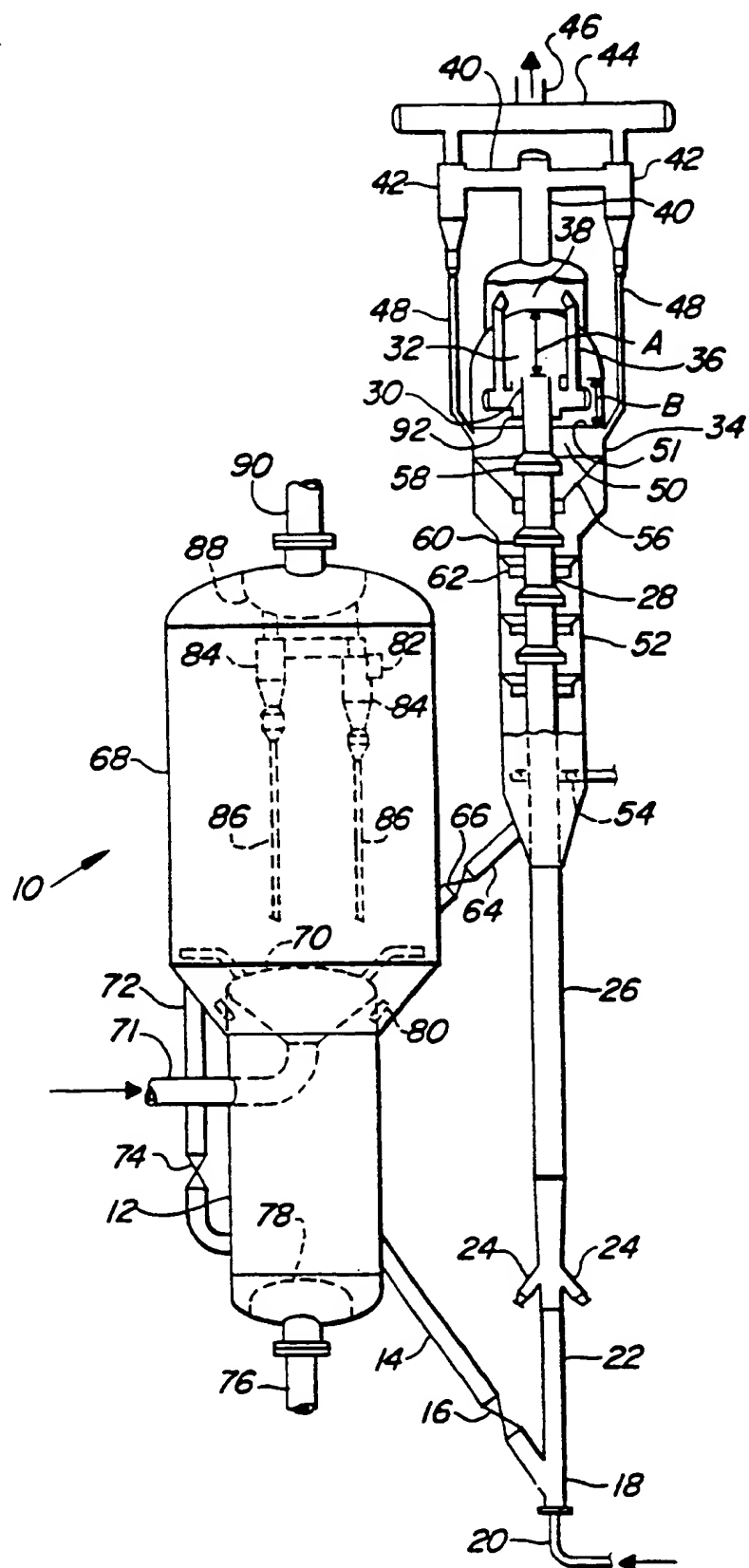
40

45

50

55

10





European Patent
Office

EUROPEAN SEARCH REPORT

Application Number

EP 92 10 6034

DOCUMENTS CONSIDERED TO BE RELEVANT			
Category	Citation of document with indication, where appropriate, of relevant passages	Relevant to claim	CLASSIFICATION OF THE APPLICATION (Int. Cl.5)
X	US-A-2 439 811 (KELLOGG) * figure 1 *	1-11	C10G11/18
X	US-A-3 406 112 (MOBIL OIL) * figure 1 *	1-11	
A	EP-A-0 065 274 (ASHLAND OIL) * figure 3 *	1-11	
A	US-A-3 993 556 (TEXACO)		
A	EP-A-0 073 396 (ASHLAND OIL)		
E	US-A-5 104 517 (UOP) * the whole document *	1-11	
			TECHNICAL FIELDS SEARCHED (Int. Cl.5)
			C10G
The present search report has been drawn up for all claims			
Place of search THE HAGUE		Date of completion of the search 04 DECEMBER 1992	Examiner MICHIELS P.
CATEGORY OF CITED DOCUMENTS X : particularly relevant if taken alone Y : particularly relevant if combined with another document of the same category A : technological background O : non-written disclosure P : intermediate document T : theory or principle underlying the invention E : earlier patent document, but published on, or after the filing date D : document cited in the application L : document cited for other reasons A : member of the same patent family, corresponding document			

(19)



Europäisches Patentamt

European Patent Office

Office européen des brevets



(11)

EP 0 564 678 B1

(12)

EUROPEAN PATENT SPECIFICATION

(45) Date of publication and mention
of the grant of the patent
03.12.1997 Bulletin 1997/49

(51) Int Cl⁶ **C10G 11/18**

(21) Application number: **92106034.9**

(22) Date of filing: **08.04.1992**

(54) **FCC process and apparatus having a low volume dilute phase disengagement zone in the reaction vessel**

FCC-Verfahren und Apparatur mit klein Volume verdünnter Phaseauslösezone in Reaktionsgefäß

Procédé et appareillage FCC avec zone de désengagement dans la phase diluée de volume réduite dans l'enceinte de réaction

(84) Designated Contracting States:
BE DE ES FR GB IT NL

(43) Date of publication of application:
13.10.1993 Bulletin 1993/41

(73) Proprietor: **UOP**
Des Plaines, Illinois 60017-5017 (US)

(72) Inventors:
• **Lomas, David A.**
Arlington Heights, Ill. 60005 (US)
• **Haun, Edward C.**
Glendale Heights, Ill. 60139 (US)
• **Sechrist, Paul A.**
Des Plaines, Ill. 60016 (US)

(74) Representative: **Weber, Dieter, Dr. et al**
Weber, Dieter, Dr.,
Seiffert, Klaus, Dipl.-Phys.,
Lieke, Winfried, Dr.
Postfach 61 45
65051 Wiesbaden (DE)

(56) References cited:
EP-A- 0 065 274 **EP-A- 0 073 396**
US-A- 2 439 811 **US-A- 3 406 112**
US-A- 3 993 556 **US-A- 5 104 517**

Remarks:

The file contains technical information submitted
after the application was filed and not included in this
specification

EP 0 564 678 B1

Note: Within nine months from the publication of the mention of the grant of the European patent, any person may give notice to the European Patent Office of opposition to the European patent granted. Notice of opposition shall be filed in a written reasoned statement. It shall not be deemed to have been filed until the opposition fee has been paid. (Art. 99(1) European Patent Convention).

Description

FIELD OF THE INVENTION

This invention relates generally to processes for the fluidized catalytic cracking (FCC) of heavy hydrocarbon streams such as vacuum gas oil and reduced crudes. This invention relates more specifically to a method for reacting hydrocarbons in an FCC reactor and separating reaction products from the catalyst used therein.

BACKGROUND OF THE INVENTION

The fluidized catalytic cracking of hydrocarbons is the main stay process for the production of gasoline and light hydrocarbon products from heavy hydrocarbon charge stocks such as vacuum gas oils or residual feeds. Large hydrocarbon molecules, associated with the heavy hydrocarbon feed, are cracked to break the large hydrocarbon chains thereby producing lighter hydrocarbons. These lighter hydrocarbons are recovered as product and can be used directly or further processed to raise the octane barrel yield relative to the heavy hydrocarbon feed.

The basic equipment or apparatus for the fluidized catalytic cracking of hydrocarbons has been in existence since the early 1940's. The basic components of the FCC process include a reactor, a regenerator and a catalyst stripper. The reactor includes a contact zone where the hydrocarbon feed is contacted with a particulate catalyst and a separation zone where product vapors from the cracking reaction are separated from the catalyst. Further product separation takes place in a catalyst stripper that receives catalyst from the separation zone and removes entrained hydrocarbons from the catalyst by counter-current contact with steam or another stripping medium. The FCC process is carried out by contacting the starting material whether it be vacuum gas oil, reduced crude, or another source of relatively high boiling hydrocarbons with a catalyst made up of a finely divided or particulate solid material. The catalyst is transported like a fluid by passing gas or vapor through it at sufficient velocity to produce a desired regime of fluid transport. Contact of the oil with the fluidized material catalyzes the cracking reaction. During the cracking reaction, coke will be deposited on the catalyst. Coke is comprised of hydrogen and carbon and can include other materials in trace quantities such as sulfur and metals that enter the process with the starting material. Coke interferes with the catalytic activity of the catalyst by blocking active sites on the catalyst surface where the cracking reactions take place. Catalyst is traditionally transferred from the stripper to a regenerator for purposes of removing the coke by oxidation with an oxygen-containing gas. An inventory of catalyst having a reduced coke content, relative to the catalyst in the stripper, hereinafter referred to as regenerated catalyst, is collected for return to the reaction zone. Oxidizing the

coke from the catalyst surface releases a large amount of heat, a portion of which escapes the regenerator with gaseous products of coke oxidation generally referred to as flue gas. The balance of the heat leaves the regenerator with the regenerated catalyst. The fluidized catalyst is continuously circulated from the reaction zone to the regeneration zone and then again to the reaction zone. The fluidized catalyst, as well as providing a catalytic function, acts as a vehicle for the transfer of heat from zone to zone. Catalyst exiting the reaction zone is spoken of as being spent, i.e., partially deactivated by the deposition of coke upon the catalyst. Specific details of the various contact zones, regeneration zones, and stripping zones along with arrangements for conveying the catalyst between the various zones are well known to those skilled in the art.

In order to obtain catalyst flow between the reactor and regenerator, a certain pressure balance is needed between the two vessels. Catalyst transferred to the regenerator is driven by the pressure in the upper section of the reactor vessel and the head of catalyst in the flow path from the reactor vessel to a control valve that controls the amount of catalyst entering the regenerator. Catalyst transfer from the regenerator to the riser is similarly driven by the pressure in the regenerator vessel and the head of catalyst in the flow path between the regenerator vessel and the bottom of the riser. There is usually a 4-10 psi (27.5-68.9 kPa) difference in pressure between the bottom of the riser and the top of the reactor vessel. Increases in pressure in the regenerator result in a corresponding increase in the pressure in the reactor. Any pressure increase of the regenerator relative to the reactor will be limited by the catalyst head in the flow path of the catalyst from the reactor to the regenerator. This interdependence of pressure often does not provide the most favorable pressure conditions in both the reactor and the regenerator. Low pressures are generally favored in the riser. Low pressures in the riser improve volatilization of the hydrocarbons and product yields. On the other hand, high pressures are favored in the regenerator to increase the coke burning capacity and improve the kinetics for coke burning. However, it is not possible to raise regenerator/reactor pressure differentials beyond a limited range due to the elevation of the available static catalyst head in the reactor vessel and stripper.

The rate of conversion of the feedstock within the reaction zone is controlled by regulation of the temperature of the catalyst, activity of the catalyst, quantity of the catalyst (i.e., catalyst to oil ratio) and contact time between the catalyst and feedstock. The most common method of regulating the reaction temperature is by regulating the rate of circulation of catalyst from the regeneration zone to the reaction zone which simultaneously produces a variation in the catalyst to oil ratio as the reaction temperatures change. That is, if it is desired to increase the conversion rate, an increase in the rate of flow of circulating fluid catalyst from the regenerator to

the reactor is effected

The hydrocarbon product of the FCC reaction is recovered in vapor form and transferred to product recovery facilities. These facilities normally comprise a main column for cooling the hydrocarbon vapor from the reactor and recovering a series of heavy cracked products which usually include bottom materials, cycle oil, and heavy gasoline. Lighter materials from the main column enter a concentration section for further separation into additional product streams

Improvements in the reduction of product losses and the control of regeneration temperatures have been achieved by providing multiple stages of catalyst stripping and by raising the temperature at which the catalyst particles are stripped of products and other combustible compounds. Both of these methods will increase the amount of low molecular weight products that are stripped from the catalyst and will reduce the quantity of combustible material in the regenerator. A variety of arrangements are known for providing multiple stages of stripping and heating the spent catalyst to raise the temperature of the stripping zone. With increasing frequency it is being proposed to raise the temperature of the stripping zone by mixing the spent catalyst with hot regenerated catalyst from the regeneration zone.

As the development of FCC units has advanced, temperatures within the reaction zone were gradually raised. It is now commonplace to employ temperatures of 525°C (975°F). At higher temperatures, there is generally a loss of gasoline components as these materials crack to lighter components by both catalytic and strictly thermal mechanisms. At 1025°F (525°C), it is typical to lose 1% on the potential gasoline yield due to gasoline components thermally cracking into lighter hydrocarbon gases. As temperatures increase, to say 1025°F (550°C), most feedstocks lose up to 6% or more of the gasoline yield due to thermal cracking of gasoline components.

One improvement to FCC units, that has reduced the product loss by thermal cracking, is the use of riser cracking. In riser cracking, regenerated catalyst and starting materials enter a pipe reactor and are transported upward by the expansion of the gases that result from the vaporization of the hydrocarbons, and other fluidizing mediums if present, upon contact with the hot catalyst. Riser cracking provides good initial catalyst and oil contact and also allows the time of contact between the catalyst and oil to be more closely controlled by eliminating turbulence and backmixing that can vary the catalyst residence time. An average riser cracking zone today will have a catalyst to oil contact time of 1 to 5 seconds. A number of riser designs use a lift gas as a further means of providing a uniform catalyst flow. Lift gas is used to accelerate catalyst in a first section of the riser before introduction of the feed and thereby reduces the turbulence which can vary the contact time between the catalyst and hydrocarbons.

The benefits of using lift gas to pre-accelerate and

condition regenerated catalyst in a riser type conversion zone are well known. Lift gas typically has a low concentration of heavy hydrocarbons, i.e. hydrocarbons having a molecular weight of C₃ or greater are avoided. In particular, highly reactive type species such as C₃ plus olefins are unsuitable for lift gas. Thus lift gas streams comprising steam and light hydrocarbons are generally used.

Riser cracking whether with or without the use of lift gas has provided substantial benefits to the operation of the FCC unit. These can be summarized as a short contact time in the reactor riser to control the degree of cracking that takes place in the riser and improved mixing to give a more homogeneous mixture of catalyst and feed. A more complete distribution prevents different times for the contact between the catalyst and feed over the cross-section of the riser such that some of the feed contacts the catalyst for a longer time than other portions of the feed. Both the short contact time and a more uniform average contact time for all of the feed with the catalyst has allowed overcracking to be controlled or eliminated in the reactor riser.

Unfortunately, much of what can be accomplished in the reactor riser in terms of uniformity of feed contact and controlled contact time can be lost when the catalyst is separated from the hydrocarbon vapors. As the catalyst and hydrocarbons are discharged from the riser, they must be separated. In early riser cracking operations, the output from the riser was discharged into a large vessel. This vessel serves as a disengaging chamber and is still referred to as a reactor vessel, although most of the reaction takes place in the reactor riser. The reactor vessel has a large volume. Vapors that enter the reactor vessel are well mixed in the large volume and therefore have a wide residence time distribution that results in relatively long residence times for a significant portion of the product fraction. Product fractions that encounter extended residence times can undergo additional catalytic and thermal cracking to less desirable lower molecular weight products.

In an effort to further control the contact time between catalyst and feed vapors, there has been continued investigation into the use of cyclones that are directly coupled (see U.S. Patent No. 4,737,346) to the end of the reactor riser. This direct coupling of cyclones to the riser provides a quick separation of most of the product vapors from the catalyst. Therefore, contact time for a large portion of the feed vapors can be closely controlled. One problem with directly coupling cyclones to outlet of the reactor riser is the need for a system that can handle pressure surges from the riser—see U.S. Patent No. 4,624,772 for an example. These pressure surges and the resulting transient increase in catalyst loading inside the cyclones can overload the cyclones such that an unacceptable amount of fine catalyst particles are carried over with the reactor vapor into downstream separation facilities. Therefore, a number of apparatus arrangements have been proposed for direct

coupled cyclones that significantly complicate the arrangement and apparatus for the direct coupled cyclones, and either provide an arrangement where a significant amount of reactor vapor can enter the open volume of the reactor/vessel or compromise the satisfactory operation of the cyclone system by subjecting it to the possibility of temporary catalyst overloads

Although direct coupled cyclone systems can help to control contact time between catalyst and feed vapors, they will not completely eliminate the presence of hydrocarbon vapors in the open space of a reactor vessel. Product vapors are still present in this open space from the stripped hydrocarbon vapors that are removed from the catalyst and pass upwardly into an open space above the stripping zone. The amount of hydrocarbon vapors is also increased by direct coupled cyclone arrangements that allow feed vapors to enter the open space that houses the cyclones. Since the dilute phase volume of the reactor vessel remains unchanged when direct coupled cyclones are used and less hydrocarbon vapors enter the dilute phase volume from the riser, the hydrocarbon vapors that do enter the dilute phase volume will be there for much longer periods of time when a direct coupled cyclone system is used. (The terms "dense phase" and "dilute phase" catalysts as used in this application are meant to refer to the density of the catalyst in a particular zone. The term "dilute phase" generally refers to a catalyst density of less than 20 lbs/ft³ (480 kg/m³) and the term "dense phase" refers to catalyst densities above 30 lbs/ft³ (480 kg/m³). Catalyst densities in the range of 20 to 30 lbs/ft³ (320 to 480 kg/m³) can be considered either dense or dilute depending on the density of the catalyst in adjacent zones or regions.) In other words, when a direct coupled cyclone system is used, less product vapors may enter the open space of the reactor vessel, but these vapors will have a much longer residence time in the reactor vessel. As a result, any feed components left in the reactor vessel are substantially lost to overcracking.

The very low gas flow rate through the reactor vessel can also promote coke deposition on the interior of the vessel. The long residence time of heavy hydrocarbons at relatively high temperature in the upper section of the reactor vessel promotes the formation of coke. These coke deposits interfere with the function of the reactor vessel by forming thick deposits on the interior of the vessel thereby insulating and locally cooling portions of the metal shell. Such locally cooled portions promote the condensation of corrosive materials that can damage the reactor vessel. In addition, other problems are created by the large coke deposits which can, from time to time, break off in large chunks and block the flow of catalyst through the vessels or conduits.

One apparatus that has been known to promote quick separation between the catalyst and the vapors in the reactor vessels is known as a ballistic separation device (see U.S. Patent No. 4,792,437) which is also referred to as a vented riser. The structure of the vented

riser in its basic form consists of a straight portion of conduit at the end of the riser and an opening that is directed upwardly into the reactor vessel with a number of cyclone inlets surrounding the outer periphery of the riser near the open end. The apparatus functions by shooting the high momentum catalyst particles past the open end of the riser where the gas collection takes place. A quick separation between the gas and the vapors occurs due to the relatively low density of the gas which can quickly change directions and turn to enter the inlets near the periphery of the riser (see U.S. Patent No. 4,295,961) while the heavier catalyst particles continue along a straight trajectory that is imparted by the straight section of riser conduit. The vented riser has the advantage of eliminating any dead area in the reactor vessel where coke can form while providing a quick separation between the catalyst and the vapors. However, the vented riser still has the drawback of operating within a large open volume in the reactor vessel.

The problem addressed by the present invention is the need to obtain a quick separation between catalyst and product vapors leaving an FCC riser in a system that minimizes overcracking of product vapors and the carryover of fine catalyst particles with the product vapors.

BRIEF DESCRIPTION OF THE INVENTION

It is an object of this invention to reduce the hydrocarbon residence time in a reactor vessel.

It is another object of this invention to improve the operation of a vented riser in the separation of catalyst and hydrocarbon vapors.

A further object of this invention is to control the residence time in a reactor vessel section of an FCC reaction zone.

This invention is an FCC process that is arranged so that the outlet end of a reactor riser discharges into an upper portion of a reactor vessel that functions as an open disengagement zone and the discharge end of the reactor riser is located near the top of a dense catalyst bed in the reactor vessel. This arrangement is facilitated by putting the cyclones or gas solid separators on the outside of the reactor vessel. It has been discovered that when a vented riser is used in combination with externally located cyclones or separation devices the upper level of a dense catalyst bed can be maintained near the top of a reactor vessel. The high level of the dense catalyst bed relative to the riser outlet reduces the dilute phase volume in the disengagement vessel so that hydrocarbon residence time is reduced and the reactor vessel height required above the dense bed level is decreased.

The decreased height of the dilute phase in the reactor vessel offers a number of benefits. The decreased dilute phase height can shorten the reactor vessel and its overall elevation. Alternately, the overall elevation of the reactor vessel can be maintained and the additional

height can be used to maintain a longer vertical length for the dense catalyst phase. Additional catalyst stripping can be performed in the additional length of dense catalyst bed. In addition, an increased height of dense catalyst increases the pressure drop between the reactor vessel and the regenerator control valve so that higher regenerator pressures can be maintained without raising the pressure in the reactor zone. The dense bed level is also susceptible to a substantial degree of variation so that the overall residence time of hydrocarbon vapors in the reactor vessel can be adjusted. This permits the use of very short residence time for certain feedstocks and an increase in residence time for more refractory feedstocks without a variation in the feed rate to the reactor riser. These benefits show that this invention will provide much of the same improvement offered by direct coupled cyclones in regard to reducing over-cracking while giving a much greater degree of flexibility that is combined with the increased reliability of an open discharge type reactor riser.

Accordingly in one embodiment, this invention is a process for the fluidized catalytic cracking of an FCC feedstock. The process includes the steps of passing the FCC feedstock and catalyst particles to a reactor riser having an effective riser outlet diameter, transporting the catalyst and feedstock upwardly through the riser thereby converting the feedstock to a gaseous product stream, and producing spent catalyst particles by the deposition of coke. The mixture of spent catalyst particles and gaseous products are discharged into a reactor vessel in an upward direction from a discharge end of the riser located from 1 to 12 outlet riser diameters below the upper end of the reactor vessel thereby providing an initial separation of the spent catalyst from the gaseous products. The separated catalyst passes downward through the vessel where it is maintained as a dense catalyst bed to form a stripping zone while a stripping gas passes upwardly through the reactor vessel. The upper surface of the bed is a minimum distance of 2 feet (.6 m) or 1 riser diameter below the riser outlet end, and a maximum distance of from 8 feet (2.4 m) to 4 riser diameters below the outlet end of the riser. A mixture of gaseous product stripping fluid and spent catalyst particles is withdrawn from the reactor vessel and transferred to a particle separator that is located outside of the reactor vessel wherein the gaseous components are separated from the spent catalyst which is ultimately returned to the regeneration zone. Spent catalyst particles are passed from the stripping zone into a regeneration zone and contacted therein with a regeneration gas in the regeneration zone to combust coke from the catalyst particles and produce regenerated catalyst particles for transfer to the reactor riser.

This invention can also be described in the context of an apparatus. The apparatus includes an upwardly directed riser conduit having an upwardly directed outlet end having an outlet opening corresponding to an equivalent diameter, referred to herein as the "riser outlet di-

ameter", a reactor vessel that surrounds the outlet end and has an upper end located 5 to 12 riser outlet diameters above the outlet end of the riser. Gas solids separation devices are located outside of the reactor vessel and these devices have an inlet, a gas outlet and a solids outlet. Means are provided for collecting a mixture of gas and catalyst from the upper half of the reactor vessel and communicating the mixture of catalyst and gas to the inlet of the separation device. Means are provided for returning catalyst particles from the collector to the reactor vessel, and means are provided for withdrawing catalyst from the bottom of the reactor vessel and transferring the catalyst to a regeneration vessel.

BRIEF DESCRIPTION OF THE DRAWINGS

The Figure shows a reactor/regenerator system for an FCC process arranged in accordance with this invention.

DETAILED DESCRIPTION OF THE INVENTION

This invention relates generally to the reactor side of the FCC process. This invention will be useful for most FCC processes that are used to crack light or heavy FCC feedstocks. The process and apparatus aspects of this invention can be used to modify the operation and arrangement of existing FCC units or in the design of newly constructed FCC units.

This invention uses the same general elements of many FCC units. A reactor riser provides the primary reaction zone. A reactor vessel and cyclone separators remove catalyst particles from the gaseous product vapors. A stripping zone removes a large percentage of sorbed catalyst particles from the surface of the catalyst. Spent catalyst from the stripping zone is regenerated in a regeneration zone having one or more stages of regeneration. Regenerated catalyst from the regeneration zone is used in the reactor riser. A number of different arrangements can be used for the elements of the reactor and regenerator sections. The description herein of specific reactor and regenerator components is not meant to limit this invention to those details except as specifically set forth in the claims.

An overview of the basic process operation can be best understood with reference to the drawing. Regenerated catalyst from a lower portion 12 of a regeneration vessel 10 is transferred by a conduit 14, at a rate regulated by a control valve 16, to a Y-section 18. Lift gas injected into the bottom of Y-section 18, by a conduit 20, carries the catalyst upward through a lower riser section 22. Feed is injected into the riser above lower riser section 22 by feed injection nozzles 24.

The mixture of feed, catalyst and lift gas travels up an intermediate section of the riser 26 and into an upper internal riser section 28 that terminates in an upwardly directed outlet end 30 that is located in a dilute phase region 32 of a reactor vessel 34. The gas and catalyst

are separated in dilute phase section 32. Conduits 36 collect gas from the dilute phase section and transfer it to a collection chamber 38. From collection chamber 38, a T-type piping arrangement 40 distributes the gas which still contains a small amount of catalyst particles to a pair of cyclone separators 42. Relatively clean product vapors are recovered from the outlets of cyclones 42 by a manifold 44 and withdrawn from the process through an outlet 46. Catalyst separated by cyclone separators 42 is carried back to reactor vessel 34 by dip pipe conduits 48. Spent catalyst from dilute phase section 32 and the dip pipe conduits form a dense catalyst bed 50 in a lower portion of the reactor vessel 34. Separated catalyst passes downwardly through the vessel at an average rate of less than 20 lb/ft²/sec (98 kg/m²/sec). The dense catalyst bed extends downward into a stripping vessel 52 that operates as a stripping zone. Stripping fluid enters a lower portion of the stripping vessel 52 through a distributor 54 and travels upward through the stripping vessel and reactor vessel in countercurrent flow to the downward moving catalyst. As the catalyst moves downward, it passes over reactor stripping baffles 56 and 58 and stripper baffles 60 and 62 and is transferred into an upper section 68 of the regenerator vessel 10 by a conduit 64 at a rate regulated by a valve 66 control. The catalyst particles are contacted with an oxygen-containing gas in upper section 68 of the regeneration zone. A distributor 70 receives the oxygen-containing gas from a conduit 71 and distributes the gas over the cross-section of the regeneration vessel. Regenerated catalyst is withdrawn from upper section 68 and transferred by a conduit 72 into lower portion 12 of the regeneration vessel at a rate regulated by a control valve 74. Catalyst in the lower portion 12 is contacted with additional regeneration gas that enters the vessel through conduit 76 and is distributed over the cross-section of the vessel by dome style distributor 78. Catalyst in lower section 12 is fully regenerated and withdrawn by conduit 14 in the manner previously described. The products of coke combustion in the lower regeneration section 12 rise upwardly and flow into the upper regenerator section 68 through a series of internal vents 80. Flue gas from lower section 12 is mixed with flue gas generated in upper section 68 and withdrawn through an inlet 82 of cyclones 84. The flue gas entering cyclone 84 contains a small amount of fine catalyst particles that are removed by the cyclones and returned to the regenerator by dip legs 86. Flue gas leaving the cyclones is collected in a chamber 88 that leaves the regenerator through a conduit 90.

The reactor riser of this invention is laid out to perform an initial separation between the catalyst and gaseous components in the riser. The term "gaseous components" includes lift gas, product gases and vapors, and unconverted feed components. The drawing shows this invention being used with a riser arrangement having a lift gas zone 22. It is not necessary that a lift gas zone be provided in the riser in order to enjoy the ben-

efits of this invention. However, the end of the riser must terminate with one or more upwardly directed openings that discharge the catalyst and gaseous mixture in an upward direction into a dilute phase section of the reactor vessel. The open end of the riser can be of an ordinary vented riser design as described in the prior art patents of this application or of any other configuration that provides a substantial separation of catalyst from gaseous material in the dilute phase section of the reactor vessel. It is believed to be important that the catalyst is discharged in an upward direction in order to minimize the distance between the outlet end of the riser and the top of the dense phase catalyst bed in the reactor vessel. The flow regime within the riser will influence the separation at the end of the riser. Typically, the catalyst circulation rate through the riser and the input of feed and any lift gas that enters the riser will produce a flowing density of between 3 lbs/ft³ to 20 lbs/ft³ (48 to 320 kg/m³) and an average velocity of about 10 ft/sec to 100 ft/sec (3 to 30 m/sec) for the catalyst and gaseous mixture. The length of the riser will usually be set to provide a residence time of between .5 to 5 seconds at these average flow velocity conditions.

The velocity at which the catalyst and gaseous mixtures discharge from end 30 of the riser also influences the placement of the end of the riser relative to the top of the reactor vessel. This distance indicated by the letter "A" in the drawing is set on the basis of the flow rate to riser. In the interest of minimizing the dilute volume of catalyst in the reactor vessel, distance "A" should be kept as short as possible. Nevertheless, there is need for some space between the end of the riser to avoid direct impingement and the resulting erosion of the top of the reactor vessel and to allow the discharge of catalyst from the end of the riser to provide a separation while preventing the reentrainment of catalyst particles that are separated by the initial discharge from the riser with the gas stream that is collected from the upper section of the reactor vessel. Since the reactor riser is usually designed for a narrow range of exit velocities between 20 to 100 ft/sec (6.1 to 30.5 m/sec), distance "A" can be set on the basis of riser outlet diameter. In order to avoid erosion of the upper surface of the reactor vessel and to promote the initial separation of the catalyst from the gaseous components, the distance "A" should be at least one outlet diameter or more. The avoidance of catalyst re-entrainment after discharge of the riser is influenced by both the riser velocity and the flowing density of the catalyst as it passes downwardly through the reactor vessel. For most practical ranges of catalyst density in the riser, the distance of 1.5 to 12 riser outlet diameters, and more preferably 1.5 to 6 riser diameters, for dimension "A" is adequate for a flowing catalyst density, often referred to as "catalyst flux", of about 50 - 200 lb/ft²/sec. (244 - 976 kg/m²/sec).

The total dilute phase volume in the reactor vessel is determined by the diameter of the reactor vessel, the distance from the end of the riser to the top of the reactor

vessel, dimension "A", and the distance from the discharge end of the riser to the top of the dense bed level in the reactor vessel which is shown as dimension "B" in the Figure. In order to prevent re-entrainment of catalyst particles into the gases that are withdrawn from the reactor vessel, a minimum distance is required from the top of the reactor riser to the top of the catalyst bed level. This dimension is primarily influenced by the superficial velocity of the gases that flow upwardly through dense bed 50. In order to minimize the potential for re-entrainment of the gaseous compounds passing through bed 50, the superficial velocity is typically below 0.5 ft/sec (.15 m/sec). The gaseous components passing upward through bed 50 are made up of stripping fluid and hydrocarbons that are desorbed from the surface of the catalyst. The amount of stripping gas that enters the stripping vessel is usually proportional to the volume of voids in the catalyst. For most reasonable catalyst to oil ratios in the riser, the amount of stripping gas that must be added to displace the void volume of the catalyst will not exceed 6 wt% of the feed rate. Accordingly, the relative ratio of catalyst passing downwardly through the reactor vessel and the stripping fluid as well as other displaced hydrocarbons flowing upwardly through the reactor vessel will remain relatively constant. Thus, the primary variable in controlling the superficial gas velocity upward through the dense catalyst bed is the diameter of the reactor vessel. As long as the superficial velocity of the gases rising through bed 50 are kept in a range below 0.5 ft/sec (.15 m/sec) and preferably below about 0.1 ft/sec (.03 m/sec), a distance "B" of 2 to 8 feet (.6 to 2.4 m) will prevent re-entrainment of the catalyst and the gases that are leaving the reactor vessel. For most reactor risers, the 2 to 8 feet will equal approximately 1 to 4 riser outlet diameters.

The manner in which the gaseous vapors are withdrawn from the dilute phase volume of the reactor vessel will also influence the initial separation and the degree of re-entrainment that is obtained in the reactor vessel. In order to improve this disengagement and avoid re-entrainment, the Figure shows the use of an annular collector 92 that surrounds the end 30 of the riser. Collector 92 is supported from the top of the reactor vessel 34 by withdrawal conduits 36. Withdrawal conduits 36 are symmetrically spaced around the annular collector and communicate with the annular collector through a number of symmetrically spaced openings to obtain a balanced withdrawal of gaseous components around the entire circumference of the reactor riser. All of the stripping gas and gaseous components from the reactor riser are withdrawn by annular collector 92 for the process arrangement shown in the Figure. All of the product gases from conduits 36 are transferred to the cyclones 42.

The Figure shows an arrangement for transferring gases from the conduits 36 to the cyclones that avoids a mal-distribution of the catalyst and gas mixture to the different cyclones. The simplest way to connect the gas

conduits with the cyclones is to directly couple one conduit to a corresponding cyclone. This arrangement would also have the advantage of minimizing the flow path between the annular collector of the riser and the cyclones where the final separation of catalyst and gas is performed. However, for reasons unknown, it has been found that mixtures of catalyst and gas that are taken from the reactor through a series of conduits may preferentially flow to one conduit. The resulting heavier loading of catalyst and gas can overload the cyclone to which it is directed. For this reason, the Figure shows the use of a chamber 38 that commonly collects the gas from all cyclone conduits 36 and redistributes the gas to the individual cyclones. Although providing chamber 38 and T-section 40 increases the residence time for the catalyst and gas mixture as it flows from the reactor vessel to the cyclone inlets, this minor increase in residence time will not have a substantial impact on the quality of the product recovered from the cyclones. The avoidance of mal-distribution may also be accomplished by the use of a catalyst and gas separation device other than cyclones.

Catalyst recovered by the cyclones can be returned to the process at any convenient location. Whatever type of gas and catalyst separation device is utilized, the catalyst separated therefrom is returned to the process. The catalyst may be returned to any point of the process that puts it back into the circulating inventory of catalyst. The drawing shows the use of conventional cyclones with the dip legs 48 returning near the upper bed level 51 of dense bed region 50. Preferably, the catalyst will be returned to the dense bed in the reactor vessel or stripping vessel.

With the cyclones removed from the reactor vessel, the diameter of the reactor vessel is no longer affected by the need to provide adequate space for a separation device therein. Accordingly, the diameter of the reactor vessel can be set on the basis of the superficial velocity of gas passing upward through the dense bed and the catalyst flux entering the reactor vessel. The criteria for both of these parameters, as previously discussed, will permit the use of a smaller reactor diameter than has been found in the prior art. The smaller reactor vessel diameter further decreases the volume of the dilute phase in the reactor vessel. When this invention is used with a new reactor vessel, the diameter can be kept low enough such that the average residence time in the dilute phase of the reactor vessel will be less than three seconds. Again, since the superficial velocity and catalyst flux are influenced by a well-known range of catalyst density and velocity conditions in the riser, the diameter of the reactor vessel when initially designed in accordance with this invention will preferably be between three and five times the outlet diameter of the riser. Alternately the dilute phase volume of the reactor vessel above the top of the catalyst bed can be kept to less than five times the volume of the reactor riser through which the feed passes.

Catalyst that is initially separated from the gaseous components as it enters the reactor vessel, passes downwardly through the vessels as previously described. As this catalyst progresses through the vessel, it preferably contacts a series of baffles that improve the contact of the catalyst with a stripping gas that passes upwardly through the vessel. In the embodiment of the invention shown in the Figure, the catalyst passes through a stripping section in the upper portion of the vessel and a separate stripping vessel located therebelow. The Figure shows the baffles 56 and 62 located on the exterior of the vessel walls and baffles 58 and 60 located down the length of the riser through the lower portion of the reactor vessel and the stripping vessel. These stripping baffles function in the usual manner to cascade catalyst from side to side as it passes through the vessel and increase the contact of the catalyst particles with the stripping steam as it passes upward in countercurrent contact with the catalyst. Dense bed 50 has a relatively long length in reactor vessel 34. There is no requirement for a long dense bed length in the reactor vessel and the dense bed length shown in the Figure stems from the type of arrangement depicted in the Figure. The Figure depicts a retrofit of this invention into an existing regeneration and reactor section where the tangent length of the reactor vessel was set by the previous arrangement that place the cyclone separators inside the reactor vessel. When the method of this invention is employed in the initial design of a reactor vessel, the tangent length can be substantially reduced so that upper bed level 51 is near the top of a stripper vessel.

Nevertheless, the height of the dense catalyst bed in the reactor 34 increases the total height of the dense phase catalyst above control valve 66. This additional height of dense bed catalyst can be used advantageously in the FCC process. First, the additional length of dense bed catalyst provides an elongated region for increased contact between the stripping fluid and the catalyst. Therefore, a greater degree of stripping can be obtained by the extended length of the dense catalyst bed. In addition, the hydrostatic head of catalyst from the top surface 51 to control valve 66 produces a relatively high pressure drop between control valve 66 and bed level 51. This pressure drop can total 7 psi (48 kPa) or more. This additional pressure allows the regenerator to be operated at a higher pressure than the reactor section. As previously described, there are substantial benefits to operating the regeneration zone at higher pressures and the reaction zone at lower pressures.

The additional height of dense bed can also be used to incorporate a hot stripping section. The hot stripping section will utilize catalyst from the regeneration zone to supply heat to the stripping section and increase the desorption of hydrocarbons and volatile components from the surface of the catalyst. A suitable lift system can be used to transport the catalyst upward from the regeneration zone into a stripping zone at a desired elevation.

The catalyst is withdrawn from the stripping zone and transferred to a regeneration zone. Any well-known regenerator arrangement for removing coke from the catalyst particles by the oxidative combustion of coke and returning catalyst particles to the reactor riser can be used.

Claims

1. A process for the fluidized catalytic cracking (FCC) of an FCC feedstock, said process comprising:
 - a) passing said FCC feedstock and catalyst particles to a reactor riser having a riser outlet diameter and transporting said catalyst and feedstock upwardly through said riser thereby converting said feedstock to a gaseous products and producing spent catalyst particles by the deposition of coke on said catalyst particles;
 - b) discharging a first mixture of spent catalyst particles and gaseous products in an upward direction into a reactor vessel from a discharge end of said riser located from 5 to 12 riser outlet diameters below the upper end of said reactor vessel thereby providing an initial separation of the spent catalyst from the gaseous products;
 - c) passing separated catalyst downward through said reactor vessel;
 - d) maintaining said separated catalyst in said reactor vessel as a dense catalyst bed to form a stripping zone and passing a stripping gas upward through the stripping zone;
 - e) maintaining the upper surface of said dense catalyst bed a distance of between one to four riser outlet diameters below said riser discharge;
 - f) passing stripped catalyst from said stripping zone into a regeneration zone and contacting said spent catalyst with a regeneration gas in said regeneration zone to combust coke from said catalyst particles and produce regenerated catalyst particles for transfer to said reactor riser; and,
 - g) withdrawing a second mixture of gaseous products, stripping fluid, and spent catalyst particles from said reactor vessel and transferring said mixture to a particle separator located outside of the reactor vessel and separating gaseous components from said spent catalyst particles in said separator zone.

2. The process of claim 1 wherein said first mixture is discharged from said riser at a velocity in a range of from 20 to 100 ft/sec (6.1 to 30.5 m/sec)
3. The process of claim 1 wherein said second gas mixture is withdrawn from an annular collector that surrounds said outlet end of said riser and said particle separators comprise at least one cyclone.
4. The process of claim 1 wherein the gaseous products discharged in step (b) have an average residence time of less than three seconds in said reactor vessel.
5. The process of claim 1 wherein said reactor vessel has a diameter that is between three and five times the outlet diameter of said riser.
6. The process of claim 1 wherein the pressure drop in said reactor vessel from the bottom to the top of said dense bed is at least 7 psi (48.3 kPa).
7. The process of Claim 1 wherein separated catalyst passes downward through said reactor vessel at an average rate of less than 20 lb/ft²-sec (98 kg/m²/sec) and the stripping gas passes upwardly through the reactor vessel at an average superficial velocity of less than 5.5 ft/sec (1.5 m/sec);
8. An apparatus for the fluidized catalytic cracking of an (FCC) feedstock, said apparatus comprising:
 - a) an upwardly directed riser conduit having an upwardly directed outlet end having an equivalent outlet diameter;
 - b) a reactor vessel surrounding said outlet end and having an upper end located 1 to 12 riser outlet diameters above said outlet end;
 - c) a gas solids separation device located outside of said reactor vessel having an inlet, a gas outlet, and a solids outlet;
 - d) means for collecting a mixture of gas and catalyst from an upper portion of said reactor vessel and communicating said mixture of catalyst and gas to the inlet of said separation device;
 - e) means for introducing a gaseous stripping medium into the reactor vessel below the outlet end; and
 - f) means for returning catalyst particles from said collector and from the bottom of the disengaging vessel to said reactor vessel.
9. The apparatus of claim 8 wherein said gas solid

separation device comprises at least one cyclone separator located outside of said reactor vessel and said means for collecting a mixture of gas and catalyst includes an annular collector that surrounds the outlet end of said riser

10. The apparatus of claim 8 wherein a dip leg communicates catalyst from each catalyst outlet to said reactor vessel and the outlet of each dip leg is located below said riser outlet
11. The process of claim 1 wherein the dilute phase volume of said disengaging zone above the top of said catalyst bed is less than 5 times the volume of said reactor riser between the point where the feed enters the riser and said discharge end.

Patentansprüche

1. Verfahren für das fluidisierte katalytische Cracken (FCC) einer FCC-Beschickung, bei dem man:
 - a) die FCC-Beschickung und Katalysatorteilchen zu einem Reaktorsteigrohr mit einem Steigrohrauslaßdurchmesser führt und den Katalysator und die Beschickung aufwärts durch das Steigrohr transportiert und dabei die Beschickung in gasförmige Produkte umwandelt und durch die Ablagerung von Koks auf den Katalysatorteilchen verbrauchte Katalysatorteilchen erzeugt,
 - b) ein erstes Gemisch von verbrauchten Katalysatorteilchen und gasförmigen Produkten in einer Aufwärtsrichtung in einen Reaktorbehälter von einem Austragende des Steigrohres, das 5 bis 12 Steigrohrauslaßdurchmesser unterhalb des oberen Endes des Reaktorbehälters liegt, austrägt und dabei eine anfängliche Trennung des verbrauchten Katalysators von den gasförmigen Produkten bewirkt,
 - c) abgetrennten Katalysator abwärts durch den Reaktorbehälter führt,
 - d) den abgetrennten Katalysator in dem Reaktorbehälter als ein dichtes Katalysatorbett hält, um eine Ausstreifzone zu bilden, und ein Ausstreifgas aufwärts durch die Ausstreifzone führt,
 - e) die obere Oberfläche des dichten Katalysatorbettes in einem Abstand zwischen einem und vier Steigrohrauslaßdurchmessern unterhalb des Steigrohraustrages hält.
 - f) ausgestreift Katalysator aus der Ausstreifzone in eine Regenerierzone führt und diesen verbrauchten Katalysator in der Regenerierzone mit einem Regeneriergas in Berührung bringt, um Koks von den Katalysatorteilchen abzubrennen und regenerierte Katalysatorteil-

- chen für eine Überführung zu dem Reaktorsteigrohr zu erzeugen, und g) ein zweites Gemisch von gasförmigen Produkten, Ausstreiffluid und verbrauchten Katalysatorteilchen aus dem Reaktorbehälter abzieht und dieses Gemisch zu einer außerhalb des Reaktorbehälters liegenden Teilchenabtrenneinrichtung überführt und gasförmige Komponente von den verbrauchten Katalysatorteilchen in der Trennzone abtrennt
2. Verfahren nach Anspruch 1, bei dem man das erste Gemisch aus dem Steigrohr mit einer Geschwindigkeit im Bereich von 20 bis 100 ft/sec (6,1 bis 30,5 m/sec) austrägt.
 3. Verfahren nach Anspruch 1, bei dem man das zweite Gasgemisch aus einem ringförmigen Sammler abzieht, der das Auslaßende des Steigrohres umgibt, und daß die Teilchentrennvorrichtungen wenigstens einen Zyklon umfassen.
 4. Verfahren nach Anspruch 1, bei dem die in Stufe (b) ausgetragenen gasförmigen Produkte eine mittlere Verweilzeit von weniger als 3 Sekunden in dem Reaktorbehälter haben.
 5. Verfahren nach Anspruch 1, bei dem der Reaktorbehälter einen Durchmesser hat, der zwischen dem Dreifachen und Fünffachen des Auslaßdurchmessers des Steigrohres liegt.
 6. Verfahren nach Anspruch 1, bei dem der Druckabfall in dem Reaktorbehälter vom Boden bis zum oberen Ende des dichten Bettes wenigstens 7 psi (48,3 kPa) beträgt.
 7. Verfahren nach Anspruch 1, bei dem der abgetrennte Katalysator abwärts durch den Reaktorbehälter mit einer durchschnittlichen Geschwindigkeit von weniger als 20 lb/ft²-sec (98 kg/m²/sec) geht und das Ausstreifgas aufwärts durch den Reaktorbehälter mit einer durchschnittlichen Oberflächengeschwindigkeit von weniger als 0,5 ft/sec (0,15 m/sec) geht.
 8. Vorrichtung für das fluidisierte katalytische Cracken einer (FCC)-Beschickung mit
 - a) einer aufwärts gerichteten Steigrohrleitung mit einem nach oben gerichteten Auslaßende mit einem äquivalenten Auslaßdurchmesser,
 - b) einem Reaktorbehälter, der dieses Auslaßende umgibt und ein 1 bis 12 Steigrohrauslaßdurchmesser oberhalb des Auslaßendes liegendes oberes Ende hat,
 - c) einer Gas-Feststoff-Trennvorrichtung, die außerhalb des Reaktorbehälters angeordnet
 - ist und einen Einlaß, einen Gasauslaß und einen Feststoffauslaß besitzt,
 - d) einer Einrichtung zum Sammeln eines Gemisches von Gas und Katalysator von einem oberen Abschnitt des Reaktorbehälters, die das Gemisch von Katalysator und Gas mit dem Einlaß der Trennvorrichtung verbindet,
 - e) einer Einrichtung zum Einführen eines gasförmigen Ausstreifmediums in den Reaktorbehälter unterhalb des Auslaßendes und
 - f) einer Einrichtung zur Rückführung von Katalysatorteilchen von der Sammeleinrichtung und vom Boden des Trennbehälters zu dem Reaktorbehälter
 9. Vorrichtung nach Anspruch 8, bei der die Gas-Feststoff-Trennvorrichtung wenigstens eine außerhalb des Reaktorbehälters liegende Zyklontrennvorrichtung umfaßt und die Einrichtung zum Sammeln eines Gemisches von Gas und Katalysator eine ringförmige Sammeleinrichtung einschließt, die das Auslaßende des Steigrohres umgibt.
 10. Vorrichtung nach Anspruch 8, bei der ein Tauchrohr Katalysator von jedem Katalysatorauslaß mit dem Reaktorbehälter verbindet und der Auslaß eines jeden Tauchrohres unterhalb des Steigrohrauslasses liegt.
 11. Verfahren nach Anspruch 1, bei dem das Volumen der verdünnten Phase der Trennzone oberhalb des oberen Endes des Katalysatorbettes geringer als das Fünffache des Volumens des Reaktorsteigrohres zwischen dem Punkt, wo die Beschickung in das Steigrohr eintritt, und dem Austragende ist.

Revendications

1. Procédé de craquage catalytique en lit fluidisé (FCC) d'une charge d'alimentation FCC, ledit procédé comprenant les étapes suivantes :
 - a) on fait passer ladite charge d'alimentation FCC et des particules de catalyseur dans la conduite montante d'un réacteur ayant un diamètre de sortie de la conduite montante et on transporte ledit catalyseur et ladite charge d'alimentation vers le haut à travers ladite conduite montante, de manière à convertir ladite charge d'alimentation en produits gazeux et à produire des particules de catalyseur épuisées par le dépôt de coke sur lesdites particules de catalyseur,
 - b) on décharge un premier mélange des particules de catalyseur épuisées et des produits gazeux vers le haut dans une cuve de réacteur depuis une extrémité de décharge de ladite colonne montante située à 5 à 12 diamètres de

sortie de la conduite montante en dessous de l'extrémité supérieure de ladite cuve de réacteur, de manière à effectuer une séparation initiale du catalyseur épuisé des produits gazeux, c) on fait passer le catalyseur séparé vers le bas à travers ladite cuve de réacteur,

d) on maintient ledit catalyseur séparé dans ladite cuve de réacteur sous la forme d'un lit de catalyseur dense pour former une zone de strippage et on fait passer un gaz de strippage vers le haut à travers la zone de strippage,

e) on maintient la surface supérieure dudit lit de catalyseur dense à une distance comprise entre un et quatre diamètres de sortie de la conduite montante en dessous de ladite décharge de la conduite montante,

f) on fait passer le catalyseur strippé de ladite zone de strippage à une zone de régénération et on met en contact ledit catalyseur épuisé avec un gaz de régénération dans ladite zone de régénération pour brûler le coke desdites particules de catalyseur et produire des particules de catalyseur régénéré qui seront transférées à ladite conduite montante du réacteur, et

g) on retire un second mélange de produits gazeux, de fluide de strippage et de particules de catalyseur épuisé de ladite cuve de réacteur, on transfère ledit mélange dans un séparateur de particules situé au-dessous de la cuve de réacteur et on sépare les composants gazeux desdites particules de catalyseur épuisé dans ladite zone de séparation.

2. Procédé selon la revendication 1, dans lequel ledit premier mélange est déchargé de ladite conduite montante à une vitesse de 6,1 à 30,5 m/sec (20 à 100 pieds/seconde).
3. Procédé selon la revendication 1, dans lequel ledit second mélange de gaz est retiré d'un collecteur annulaire qui entoure ladite extrémité de sortie de ladite conduite montante et lesdits séparateurs de particules comprennent au moins un cyclone.
4. Procédé selon la revendication 1, dans lequel les produits gazeux déchargés à l'étape (b) ont un temps de séjour moyen de moins de trois secondes dans ladite cuve de réacteur.
5. Procédé selon la revendication 1, dans lequel ladite cuve de réacteur a un diamètre qui se situe entre trois et cinq fois le diamètre de sortie de ladite conduite montante.
6. Procédé selon la revendication 1, dans lequel la chute de pression dans ladite cuve de réacteur du fond au sommet dudit lit dense est d'au moins 48,3

kPa (7 psi)

7. Procédé selon la revendication 1, dans lequel le catalyseur séparé passe de haut en bas à travers ladite cuve de réacteur à un débit moyen inférieur à 98 kg/m²/sec (20 livres/pied carré-sec) et le gaz de strippage passe de bas en haut à travers la cuve de réacteur à une vitesse de surface moyenne inférieure à 0,15 m/sec (0,5 pied/sec).

8. Appareil de craquage catalytique en lit fluidisé d'une charge d'alimentation (FCC), ledit appareil comprenant :

a) une conduite montante dirigée vers le haut, ayant une extrémité de sortie dirigée vers le haut ayant un diamètre de sortie équivalent,

b) une cuve de réacteur entourant ladite extrémité de sortie et ayant une extrémité supérieure située 1 à 12 diamètres de sortie de la conduite montante au-dessus de ladite extrémité de sortie,

c) un dispositif de séparation solides-gaz situé à l'extérieur de ladite cuve de réacteur, ayant une entrée, une sortie de gaz et une sortie de solides,

d) des moyens pour recueillir un mélange de gaz et de catalyseur d'une partie supérieure de ladite cuve de réacteur et pour mettre en communication ledit mélange de catalyseur et de gaz avec l'entrée dudit dispositif de séparation,

e) des moyens pour introduire un agent de strippage gazeux dans la cuve de réacteur en dessous de l'extrémité de sortie, et

f) des moyens pour renvoyer les particules de catalyseur dudit collecteur et du fond de la cuve de dégagement à ladite cuve de réacteur.

9. Appareil selon la revendication 8, dans lequel ledit dispositif de séparation solides-gaz comprend au moins un séparateur à cyclone situé à l'extérieur de ladite cuve de réacteur et lesdits moyens pour recueillir un mélange de gaz et de catalyseur comprennent un collecteur annulaire qui entoure l'extrémité de sortie de ladite conduite montante.

10. Appareil selon la revendication 8, dans lequel un élément de conduite immergé fait communiquer le catalyseur issu de chaque sortie de catalyseur avec ladite cuve de réacteur et la sortie de chaque élément de conduite immergé est située en dessous de ladite sortie de la conduite montante.

11. Procédé selon la revendication 1, dans lequel le volume de la phase diluée de ladite zone de dégagement au-dessus du sommet dudit lit de catalyseur est inférieur à 5 fois le volume de ladite conduite montante du réacteur entre le point où l'alimentation

pénètre dans la conduite montante et ladite extré-
mité de décharge

5

10

15

20

25

30

35

40

45

50

55

